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# SYNGAS PRODUCTION WITH A DUAL FLUIDIZED BED GASIFIER FOR POLYGENERATION

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## ABSTRACT

A pilot scale dual fluidized bed gasification system was developed for polygeneration with biomass. The gasification system is designed for supplying syngas for Fischer Tropsch (F-T) synthesis of bio-diesel and power generation with a syngas engine. Characteristics of biomass steam gasification were investigated in a lab scale bubbling fluidized bed, and hydrodynamics of a dual fluidized bed were investigated in a cold flow model. Based on the results from the lab scale test and cold flow model, a pilot scale dual fluidized bed gasifier was designed. In this paper, the developing process of the gasifier and preliminary results of system operation will be presented.

## INTRODUCTION

With the recent high rocketing price of fossil fuel as well as the global warming issues, gasification becomes a very promising technology to get alternative energy from low rank fuels. Recently, an extensive research project on biomass gasification including gas cleaning and F-T synthesis process has been launched in Korea. For the project, a dual fluidized bed gasification system is adopted for producing medium heating value gas which satisfies the requirement of polygeneration. Steam is used as a gasifying agent to control hydrogen content in the product gas. For the F-T synthesis, proper gas composition is important and extremely purified syngas is necessary for avoiding any contamination of catalyst, Boerrigter and Den (1). There have been many studies about the possibility of synthetic bio fuels via biomass gasification, Tijmensen et al (2), Srinivas et al (3), Tristantini (4).

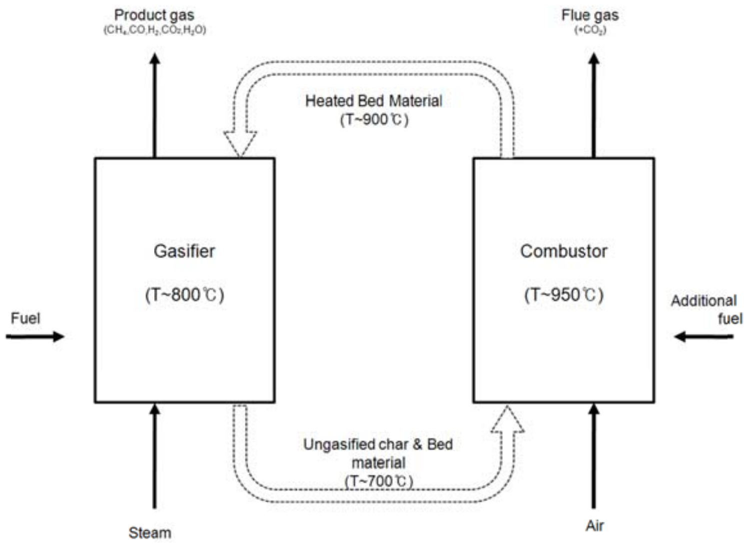



Fig. 1 Schematic diagram of a dual fluidized bed gasification system

In this paper, the development of a dual fluidized bed biomass gasification system will be presented. Results from the lab scale experiment, cold model test, and design and initial results from the pilot scale gasification system will be discussed.

## DUAL FLUIDIZED BED GASIFICATION SYSTEM

Figure 1 shows the schematic diagram of a dual fluidized bed (DFB) gasification system. In the DFB gasification system, gasifier is separated from combustor and bed material is used as a heat carrier between the two reactors. Medium heating value gas can be obtained from the gasifier because this allothermal system inherently prevents the mixing between the product and combustion gas. Recently, Corella, J et al (5) presented a review paper on DFB biomass gasifiers. According to the paper, the first application of the system on biomass gasification was conducted by Prof. D. Kunii in 1975. After that, there have been many attempts of application of DFB to biomass gasification by Kunii's group in Japan, Battelle-Columbus and FERCO in US, TU Wien and Güssing in Austria, ECN in the Netherlands, and Chalmers in Sweden.

## LAB SCALE EXPERIMENT

In this study, Korean pine wood chips were used as a feedstock. The size of the wood chip is below than 10 mm and it contains small amount of sulfur. Figure 2

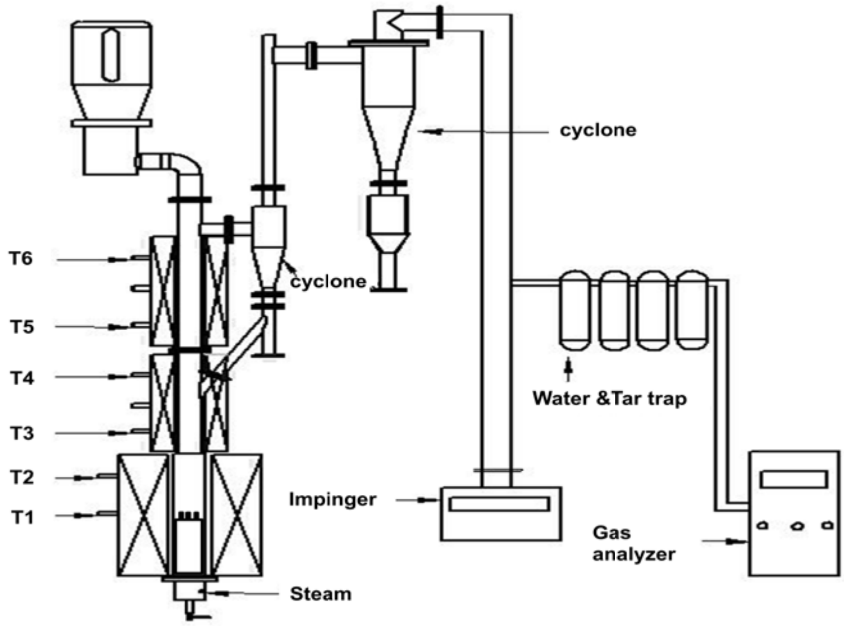



Fig. 2 Schematic diagram of the experimental set-up for the gasification (7)

Table 1 Syngas composition as a function of temperature and steam to biomass ratio

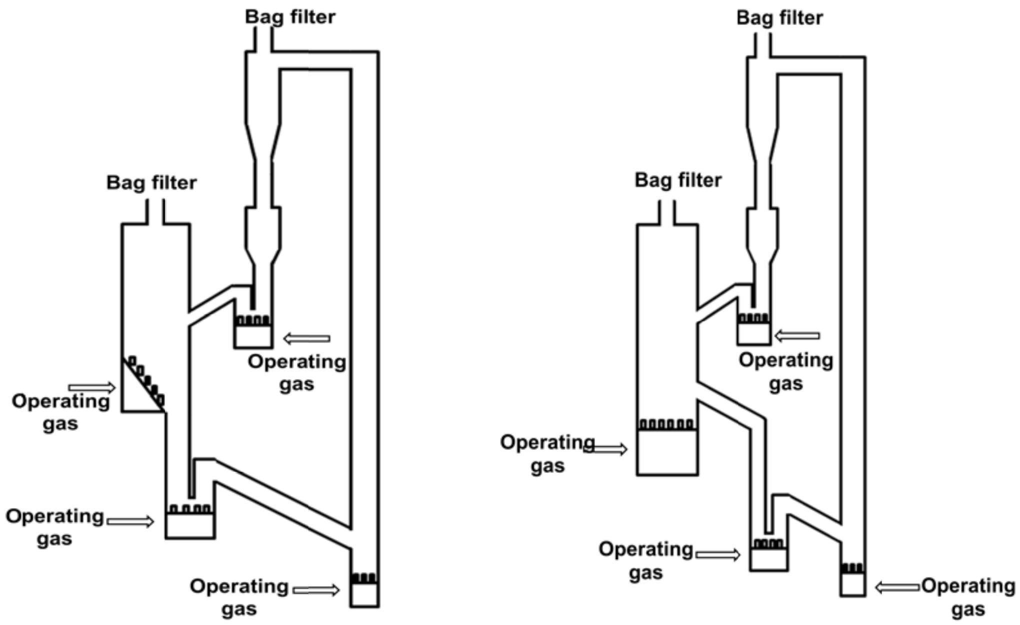
| Temperature | 700 | 800 | 900 | | 800 | |
|---|---|---|---|---|---|---|
| S/B ratio | 0.3 | 0.3 | 0.3 | 0.3 | 0.5 | 1.0 |
| $H_2$ (vol.%) | 26.3 | 34.9 | 41.9 | 34.9 | 35.1 | 33.7 |
| CO (vol.%) | 36.5 | 30.1 | 25.3 | 30.1 | 30.6 | 31.4 |
| $CO_2$ (vol.%) | 16.0 | 17.3 | 19.2 | 17.3 | 17.5 | 17.1 |
| $CH_4$ (vol.%) | 11.1 | 9.5 | 8.0 | 9.5 | 9.6 | 9.8 |
| Others (vol.%) | 10.1 | 8.2 | 5.5 | 8.2 | 7.1 | 7.9 |
| $H_2S$ (ppm) | 21.5 | 13.6 | 9.8 | 13.6 | 12.7 | 14.7 |
| COS (ppm) | 0.8 | 0.3 | 0.2 | 0.3 | 0.4 | 0.3 |
| **$H_2$/CO ratio** | **0.7** | **1.2** | **1.7** | **1.2** | **1.2** | **1.1** |

shows the lab scale experimental set up. The experimental set up is designed to investigate the gasifier part and the reactor(0.15 m I.D. × 3.0 m height) was placed inside electric furnaces. Steam was used as fluidizing/gasifying agent. The details of experimental method can be founded in the previous studies, Song et al (6), Song et al (7). The feedstock was supplied from the top of the gasifier at a rate of 5 kg/hr. Syngas compositions were monitored by an on-line gas analyzer (ND-IR analyzer, ABB CO. Ltd.) and the product gas was also collected in a gas collection bag for the investigation of sulfur compounds ($H_2S$, COS). Highly sensitive FPD(Flame Photometric Detector) and gas chromatography (Agilent 7890A) were employed for quantitative analysis of sulfur compounds.

Table 1 shows the syngas composition as a function of temperature and steam to biomass ratio (S/B). The gasification temperature was varied as 700℃, 800℃, and 900℃, and steam to biomass ratio was varied as 0.3, 0.5 and 1.0. The second column of Table 1 shows the effect of temperature when the S/B is kept as 0.3. As shown in the table, $H_2/CO$ ratio increases from 0.7 to 1.7 as gasification temperature increases from 700℃ to 900℃ which results from the reaction related to steam, Jansawang et al (8). Regarding sulfur compounds, the concentration of $H_2S$ is much higher than COS, and their contents decrease as gasification temperature increases. In this project, methanol absorption is applied to remove sulfur compounds and as well as reduction of $CO_2$. The third column of Table 1 shows the effect of steam on the product gas. In this case S/B ratio is varied from 0.3 to 1.0 while the reactor temperature is kept constant as 800℃. The results show that the amount of steam has little effect on the product gas if S/B is more than 0.3. With this lab scale experiment, we decided that the condition of $T_{gasifier}$ = 800℃ and S/B = 0.3 is suitable for the F-T synthesis because the product gas satisfies the design parameters such as $H_2/CO$ ratio and sulfur compound content before the cleaning process.

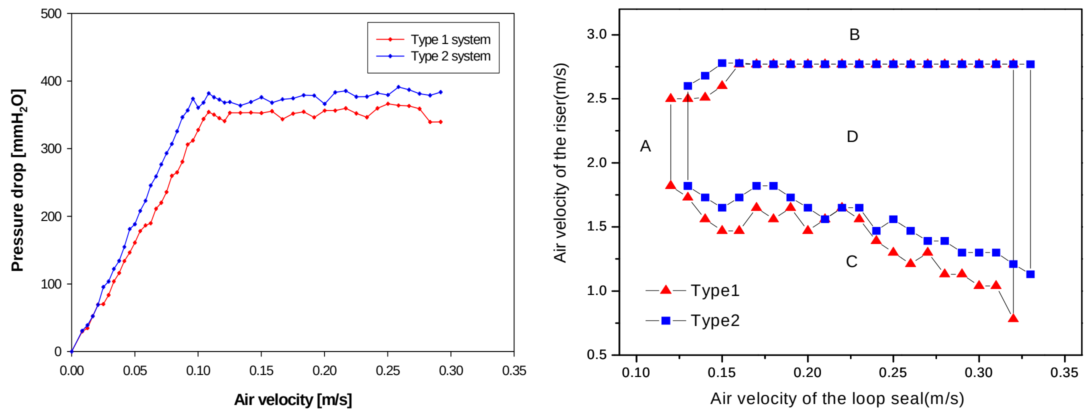
## COLD FLOW MODEL

A cold model, consisting of interconnected bubbling and fast fluidized beds, was tested for investigating the hydrodynamics of DFB system. To find an optimal circulation of heavy solid inventory between the two reactors, we investigate two types of DFB reactors which have different configuration of distributor and way-out location of the solid inventory in the gasifier part. Details of the experimental methods can be founded in the previous work, Sung et al (9). To extract appropriate design parameters and operating conditions for a pilot scale DFB gasification system, we measured minimum fluidization velocity, possible operating range, solid circulation rate, axial solid holdup and gas bypassing between the lower loop seal and the gasifier. The silica sand belonged to Geldart group B was used as a bed material. Figure 3 shows the schematics of DFB cold model consisting of a fast fluidized bed riser (0.07 m I.D × 5.18 m high) and a bubbling fluidized bed reactor (0.2 m wide × 0.2 m depth × 2.95 m high). For the two models, there are differences in the location of the solid exit and distributor of the gasifier. For Type 1, the solid exit of the gasifier is located at the bottom of the gasifier and the distributor is installed on a slat wall. For Type 2, the way-out is located on the right side of the gasifier which is upper than the distributor and the distributor is installed on a flat surface. Other than the solid exit location and the shape of distributor, all the other parts are the same for the two types.




(a) Type 1 (B) Type 2

Fig. 3 Schematics of dual fluidized bed gasification systems (9) (1: Bubbling fluidized bed, 2: Lower loop seal, 3: Riser, 4: cyclone, 5: hopper, 6: Upper loop seal)




(a) (b)

Fig. 4 Results from the cold model test: (a) minimum fluidization velocity, (b) map of stable operation as a function of air velocity of the riser and lower loop seal

Figure 4(a) shows the minimum fluidization velocity of each type. With the same silica sand, the minimum fluidization velocity of Type 1 is 0.12 m/s and that of Type 2 is 0.1 m/s which shows that minimum fluidization velocity is influenced by the geometry of the reactor and the distributor. The minimum fluidization velocity of Type 2 is closer to the calculated value: 0.09 m/s. Figure 4(b) shows the possible operating range as a function of air velocity of lower loop seal and riser. To find a stable operation region, we investigated characteristics of solid movement as a function of riser and lower loop seal velocity. As shown in the map, DFB reactor can be operated in a stable manner inside the boundary (region D). There are little difference in the stable operation map between Type 1 and Type 2 and the stable operating region of Type 1 is larger than Type 2 for all conditions except for the low velocity condition of the lower loop seal and high velocity of the riser, Sung et al (9). Regarding the minimum fluidization velocity, Type 2 is better than Type 1, but Type 1 has advantages in the view point of stable operation because the location of solid exit of Type 1. For stable operation, the bed height of Type 2 should always be kept higher than the solid exit, but in case of Type 2, the system works independent of the bed height which is robust for any changes in the operation conditions.

## PILOT SCALE DFB GASIFICATION SYSTEM

Figure 5(a) shows the schematic diagram of 1 ton/day pilot scale DFB gasification system. It consists of a bubbling fluidized bed (BFB) gasifier and circulating fluidized bed (CFB) combustor. Both reactors have a preheating chamber for initial heating and LPG is burnt with an air in the preheating chamber until the reactor temperature reaches 450 ℃ and after then main fuel is introduced for further heat up process. In the BFB gasifier, biomass is introduced from the top and solid exit is placed in the side wall of the gasifier (Type 2 of Fig. 3). In the CFB combustor, additional biomass is fed into the mid-section of the fluidized bed to control the thermal load of the combustor.

Figure 5(b) shows the temperature change of the combustor and upper loop seal as a function of time. The location of the measurement is depicted on Fig. 5(a). As shown in the figure, the temperature of the combustor increases as time increases. In the initial stage, the temperature difference between the lower and upper part of the combustor are rather large but it becomes small as the temperature of the whole system increases. About 3.5 hours later, the temperature of upper loop seal (U-1) increases rapidly, which implies the solid circulation is started at the moment because of the increased flow rate due to the increased temperature and reactant supply. It is notable that the axial temperature distribution of the combustor is significantly decreased after the solid circulation. Figure 5(c) shows the temperature change of the gasifier and lower loop seal. The bed height is just below the location of G-3. As shown in the figure, the temperature of the gasifier increases as time increases. Before the beginning of the solid transportation from the combustor, combustion gas from the chamber heats up the bed material and the gasifier.

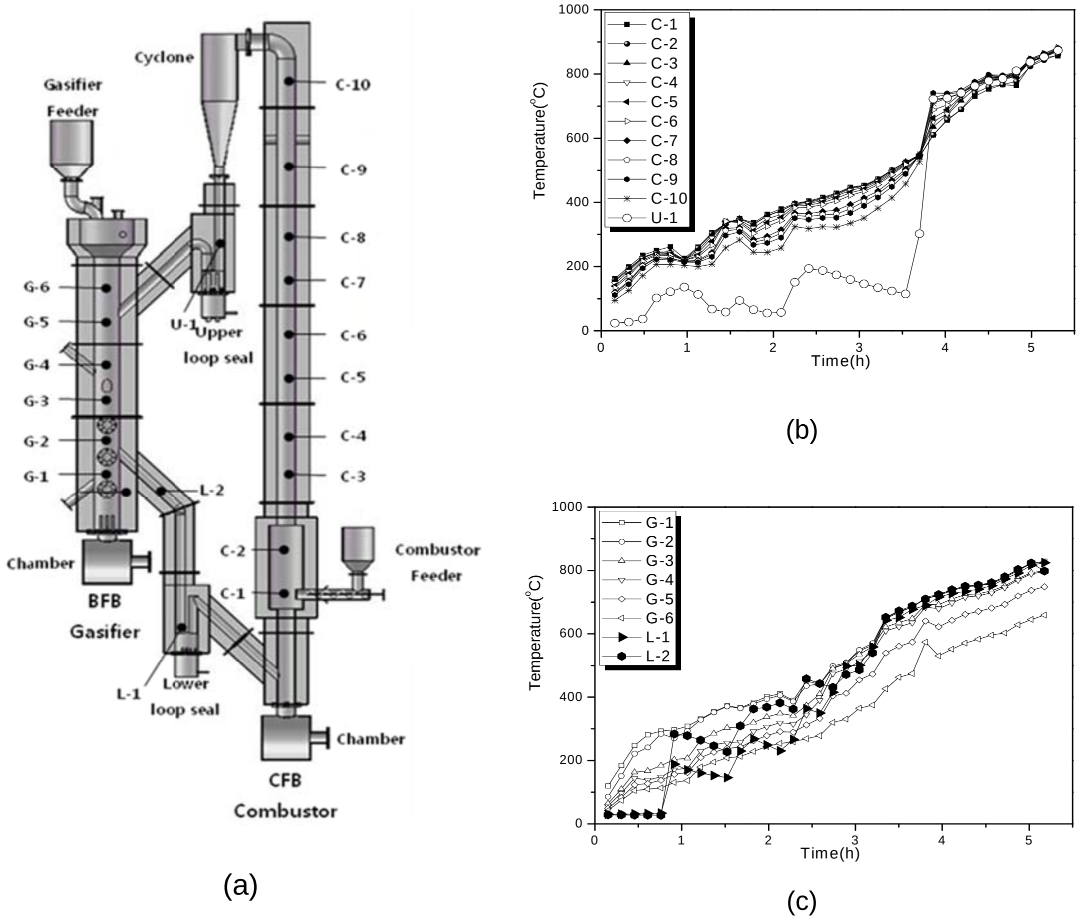



Fig. 5 (a) Schematic diagram of a pilot scale DFB gasification system, (b) Temperature change of combustor and upper loop seal (c) Temperature change of gasifier and lower loop seal

Abrupt change of the lower loop seal temperature indicates the turn on and off of the air flow into the lower loop seal. The graph shows that the temperature of inside of the bed and lower loop seal increases rapidly after the beginning of the solid circulation from the combustor.

## CONCLUSIONS

A developing process of a pilot scale dual fluidized bed gasification system was described in the paper. Lab scale experiments were conducted for finding optimum operating conditions such as reactor temperature and steam to biomass ratio. A cold model test was also conducted for two different types of gasifier configuration and

Type 2 was selected for the pilot plant design. 1 ton/day DFB gasification system was built and initial results, such as combustor and gasifier temperature change and solid circulation conditions, were obtained in the pilot plant. 1 ton/day pilot scale gasification–cleaning–polygeneration including FT synthesis system will be in operation in year 2010.

## ACKNOWLEDGEMENT

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